

Industrial wastewater desalination using electrodialysis: evaluation and plant design

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Abstract Industrial processes usually generate streams enriched with high organic and inorganic components. Due to the complexity of these streams sometimes it is not quite straightforward to predict the performance of desalination technologies. Some technologies are available for the selective removal of salts from aqueous stream, but in general these technologies are applied in high value applications where salts are either the product or limit further purification of the final product is required. These technologies are, however, not widely used in low value applications like wastewater treatment. The aim of this article is to review, improve and perform the design of electrodialysis processes for relevant industrial wastewater applications. It is focused on the determination of the critical design parameters like membrane resistance, current efficiency and limiting current density through lab scale experiments and its further use for industrial scale first approximation design. In this article, the basic equations for design are reviewed and a practical approach to

obtain the number of stacks required for a certain separation is introduced. An industrial wastewater stream has been used for lab batch experiment and its following continuous plant design. The results show that it is possible to separate monovalent ions in a high rate (more than 70 %) and divalent ions were less separated (less than 50 %). The energy required for the particular case was evaluated in a range from 6 to 11 kWh/m³ of feed stream depending on the water reclamation rate.

Keywords Electrodialysis · Waste water · Electromembrane processes · Ion-exchange membrane

1 Introduction

Aqueous-based industrial processes lead to water streams high in organic and inorganic components. The economic value of these components is, due to the matrix they are in, low or even negative. Next to this, these components often have a negative impact on the overall process sustainability. Quite some technologies are available for the selective removal of salts from aqueous streams. In general, these technologies are applied in high value applications where salts are either the product or limit further purification of the final product. These technologies are, however, only little used in low value applications like wastewater treatment where most of the desalination plants are reverse osmosis based or thermal separation processes [1]. Electrodialysis (ED) is a membrane-based separation process in which ions are driven through an ion-selective membrane under the influence of an electric field [2]. For certain applications ED has become more interesting on the last decade, like: treatment of RO concentrate [3], brackish water desalination [4], organic acids production [5],

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hardness removal [6, 7], etc. Moreover, when the polarity of the electrodes is reversed the system is called electro-dialysis reversal (EDR), the main benefit of this procedure over conventional ED or other pressure-driven membrane processes is the self-cleaning of the membranes over the long-term operation, ensuring a stable flux and desalination rate [8].

A process design for ED plants has been presented by Strathmann and co-workers [9–11] in many scientific articles. It has also been used and modified by Brauns et al. [12] in order to reduce the numbers of uncertain parameters. In general, some wastewaters can contain suspended solid materials that might interfere in the ED good performance; therefore, it is fundamental to pre-treat the influent in order to ensure the long-term operation and membrane lifetime. It is our aim to evaluate the ability of the ED system to handle suspended solid material when it is operated using thicker spacers than usual. The main aim of this article is to review, improve and perform the design of ED processes for relevant industrial wastewater applications. It focuses on the determination of the critical design parameters like membrane resistance, current efficiency and limiting current density (LCD) through lab scale experiment and its further use for industrial scale first approximation design. In this article, the basic equations for design are reviewed and a practical approach to obtain the number of stacks required for a certain separation is introduced. The design of a continuous operated ED plant is shown, treating an industrial wastewater stream, based on the theoretical model and lab batch experiments.

2 Theoretical background

In the following sections, the fundamentals for the design of an ED plant are presented. In a normal process, the stack contains the following key parts: electrodes, spacers and ion-exchange membranes arranged in the way of obtaining two or more separated streams. When the ion-exchange membranes are arranged according to Fig. 1, concentrate and dilute streams can be obtained from the brine. In general, when applying current over the stack, anions and cations are attracted towards the anode and cathode electrodes, respectively. When they reach a membrane, some will pass through and some will be rejected, depending on their charge, size and hydrophilicity effects. Water molecules will also pass through the membrane, as solvation shield in the ions. The flow channel formed between membranes is created with spacers. The main difference among the commercially available spacers is the nature of flow paths, determining the concentration polarisation layers, fouling tendency and pressure drop [13].

In conventional ED, the compartments surrounding the electrodes are normally filled with an electrolyte solution in a closed loop in order to neutralise pH and salinity. Effluents from the electrode compartments may contain gases; oxygen, hydrogen from water dissociation and other gases like chlorine depend on the source water.

2.1 LCD determination

The LCD is the equilibrium occurring in the boundary layers in the membranes due to concentration polarisation effect in the diluted side. It mainly depends on parameters like feed flow velocity, applied current density, spacers, cell design and concentration. A typical I - V curve in a wide range of voltage can have the three sections shown in Fig. 2. The first part is governed by Ohm's law with a constant resistance. When the voltage increases enough to achieve the ions depletion in the diluate polarisation layer, the concentration decreases as well, in this point any increase in the voltage lead to a very small increase in current trough the stack. When the concentration in the limiting layer reaches zero, the water splitting process starts releasing H^+ and OH^- and those ions are also been transported through the membranes and the electric resistance decreases [11]. The determination of LCD in ED is usually experimentally determined, principally because the concentration polarisation layer formation depends on several parameters which complicate the theoretical prediction [12, 14]. An accepted empirical procedure is represented in Eq. 1, relating the solution flow velocity and concentration with the LCD (i_{lim}) in the stack in an exponential function.

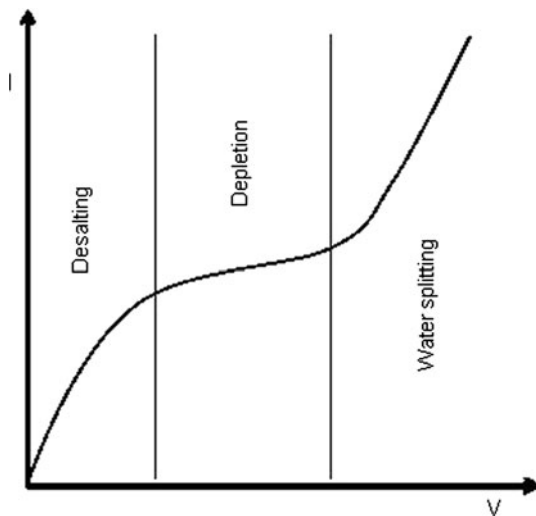
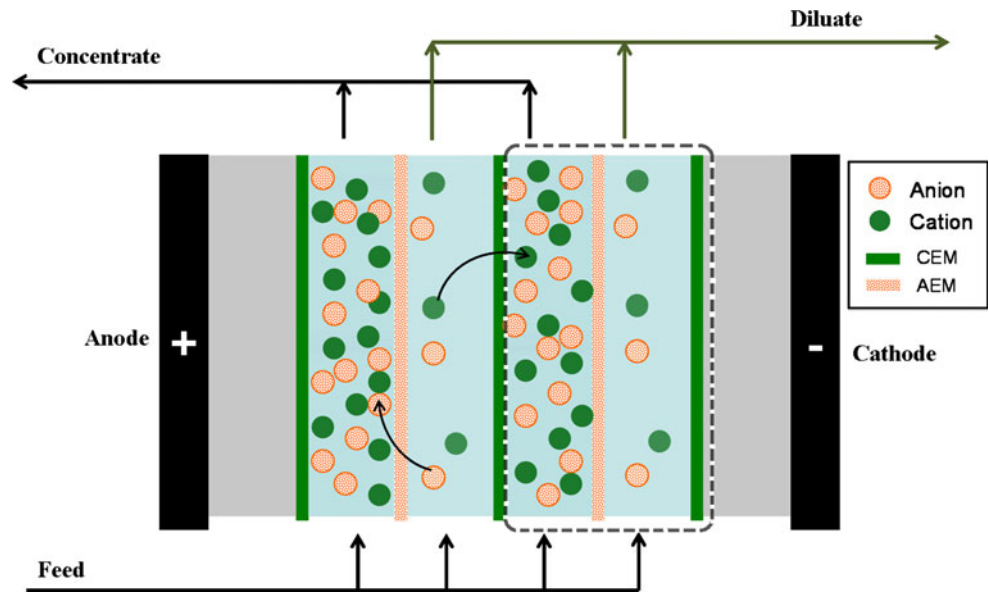
$$i_{lim} = nC^a v^b, \quad (1)$$

where C represents the ion concentration (mol/m^3) and v is the flow velocity in the stack (m/s). The coefficients in the equation, n , a and b , should be estimated by LCD measurements with different linear flow velocities and concentrations for each specific cell design. The i_{lim} values of every experiment can be obtained from the representation of resistance versus the reverse of the current ($1/i$). The i_{lim} value is on the intersection point of the slopes when $i \rightarrow 0$ and $i \rightarrow \infty$ [14].

As the LCD value is a limiting condition, in order to obtain the highest current efficiency possible, the current through the stack should be maintained below the LCD value. Therefore, in a real application the applied current should be 70–90 % of the LCD value [11].

2.2 Area determination

The design of any ED plant starts with the determination of the required membrane area for a certain capacity:

Fig. 1 Electrodialysis stack schematic representation**Fig. 2** Voltage versus current standard representation

$$A_T = \frac{\eta F}{i \xi}, \quad (2)$$

where η is the separation rate in mol/s, F is the Faraday constant, i is the applied current density in A/m² and ξ is the current efficiency. This area represents the total cell area that consists of a cation and an anion exchange membrane.

The number of required cell pairs can be obtained from Eq. 3, where H and w are the height and width of the membrane sheet, respectively.

$$N_C = \text{roundup} \left(\frac{A_T}{Hw} \right). \quad (3)$$

2.3 Current efficiency

By definition, the current efficiency is the ratio of current actually transported by a certain ion in the solution to the real (total) current passing through the stack. The current transported by ions is [15].

Number of chemical eq. of salt effectively transported.

$$N_{eq} = \frac{(n_f - n_d)}{\Delta t} \quad (4)$$

Number of electrical eq. of passed through the membrane pairs

$$I_m = NI \quad (5)$$

Current efficiency

$$\xi = \frac{(n_f - n_d)F}{NI\Delta t}, \quad (6)$$

where n_f and n_d are the equivalents transferred in a certain period Δt in a batch experiment, N is the number of cells used and I is the total current applied.

The current efficiency in an ED system is usually lowered by [9]:

- Selectivity of the membranes
- Internal leakages
- Electroosmosis (water transported in the hydration layer of the ion)
- H⁺ and OH⁻ transport (mostly at high current density and low concentration)
- Co-ion transport; usually referred as back diffusion.

From a batch experiment, it is possible to obtain the mean efficiency of the process as function of time or

concentration, which will include all the possible deviations like: electro osmosis effects, ion competition and/or internal leakages.

2.4 Resistances and cell voltage drop

According to the characteristics of the system, a voltage drop is generated as consequence of the ion current through the stack. In a conventional ED stack, there are several contributions to the global resistance. In Fig. 3, a schematic profile of concentrations is represented for a unit cell formed by one CEM and AEM:

- At the surface of an ion-exchange membrane, a gradient of concentration at both sides of the membrane is developed due to the transport of charged species through it and the cross-flow solution in the compartment. The ion movement generates two different equilibrium boundary layers facing each side (concentration polarisation). Irrespective of the charge of the membrane, there will be a concentrate and a diluate boundary layer at each side. Although, all this layers constitute a resistance by themselves, the diluate boundary layer is the point with highest resistance in the whole system assuming that the membrane itself is permeable and conductive enough to not become the limiting element

- Inside the membranes there will be a concentration profile from the highest to the lowest concentration point in the cell. The resistance of the swollen membrane is variable depending on the solution composition.
- The resistance of the bulk compartments (diluate and concentrate) is determined by the logarithmic mean value between inlet and outlet conductivity of the stack [11].

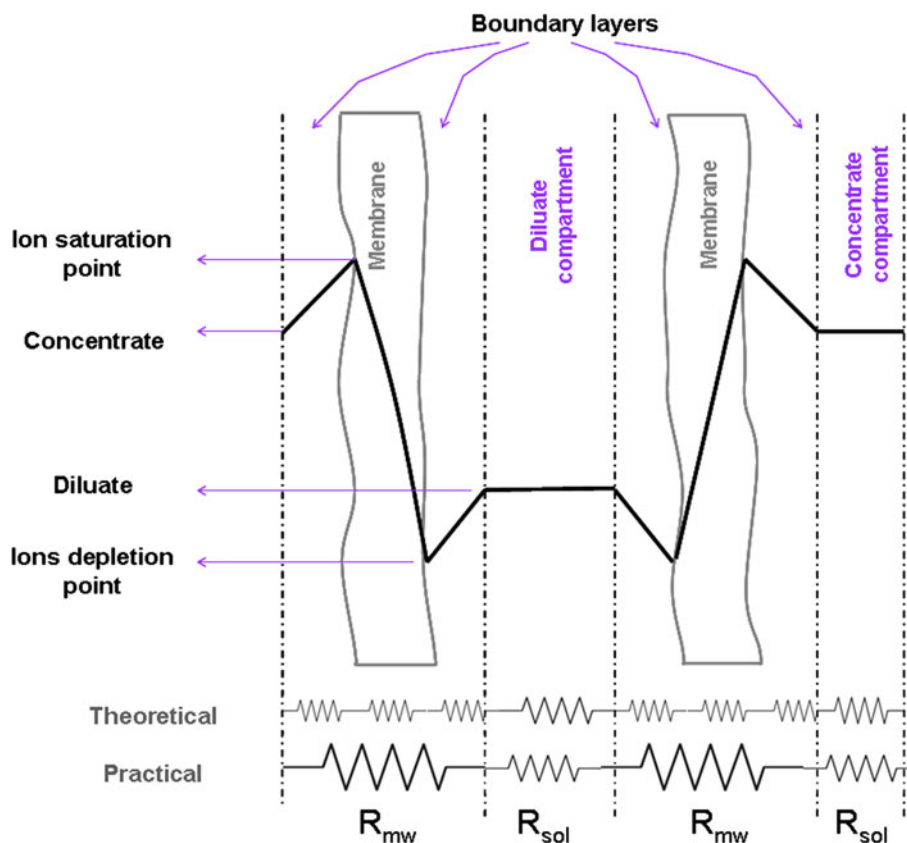
Unfortunately, the two first contributions to the overall resistance are not easy to obtain experimentally or without complex mathematical modeling. In order to predict reliable values of the cell voltage drop, the boundary layers and swollen membrane resistances are coupled in one unique resistance. Membranes suppliers usually inform resistance values that were measured under high concentration of NaCl (normally 1 or 2 M) and consequently the resistance is lower than in most of desalination applications.

Consequently, there are two ‘practical resistance’ contributions to the global voltage developed:

- (1) Bulk solution voltage:

$$\Delta U_{\text{sol}} = i \left(\sum_k \frac{h}{\lambda_k} \right) \quad (7)$$

Fig. 3 Schematic representation of the theoretical components for the cell resistance and the practical approach



(2) Membrane voltage:

$$\Delta U_{\text{mem}} = i \sum_k R_{\text{mw}}, \quad (8)$$

where λ_k is the logarithmic average conductivity in each k compartment (S/m) of h thickness and R_{mw} is the resistance of the swollen membranes in $\Omega \text{ m}^2$.

Therefore, the total cell voltage drop (ΔU_{cell}) can be calculated by adding Eqs. 7 and 8.

2.5 Energy consumption

The energy consumption for the desalination process is obtained as

$$E = i A_T \Delta U_{\text{cell}} \quad (9)$$

And the specific energy consumption is

$$\bar{E} = \frac{E\tau}{Q_0\rho} \quad (10)$$

with τ representing the amount of working hours by year, Q_0 and ρ are the flow and density of stream to treat, respectively.

For a certain current density applied the stack design is well established by Eqs. 2–10.

2.6 Variable current density application

In a real process, the current density set in each stack or stack groups could be different in order to improve the energy efficiency, starting from high values and decreasing step by step towards the final concentration. Following this concept, it is possible to calculate the required membrane area with the same equations presented in Sects. 2.1–5. This approach will help to obtain the number of stacks required.

In order to determine the optimum current density in each step, some conditions were imposed to arrange the variable current density through the length:

- Each stack will have the same amount of membrane area, consequently same number of cell pairs.
- Any stack cannot reach more than 200 cell pairs or exceed the 300 V of total voltage. These values were arbitrary selected; more cells with total voltage per stack can be arranged.

Following these two practical constraints, the current density distribution in the stacks will be set from the first stack to the last in decreasing values, according to the following description of Eqs. 11–14.

From the equal membrane area condition

$$A_1 = A_2 = \dots = A_k = \dots = A_{N-1} = A_N = \frac{Q\Delta C_k^d F}{i_k \zeta} \quad (11)$$

For two generic stacks k and $k + 1$, as shown in Fig. 4, and assuming there is not water transport through the membrane and equal current efficiency in all stacks

$$\frac{\Delta C_k^d}{i_k} = \frac{\Delta C_{k+1}^d}{i_{k+1}} \quad (12)$$

According to Eq. 1, the electric current density can be correlated with a power equation

$$i_k = n(C_{k+1}^d)^a v^b, \quad (13)$$

where a , b and n are parameters obtained experimentally. Then substituting Eq. 13 into 12

$$\frac{(C_k^d - C_{k+1}^d)}{(C_{k+1}^d)^a} = \frac{(C_{k+1}^d - C_{k+2}^d)}{(C_{k+2}^d)^a} \quad \text{for } k = 1, 2, \dots, N-1 \quad (14)$$

Equation 14 can be solved for the inter-stack concentrations in the diluate side. In that scheme the concentrate stream could be set on co-current or counter current respect to the diluate. Then the membrane area and current density can be obtained using Eqs. 11 and 13, respectively. The number of stacks required for a certain separation will be assessed using the algorithm resolution shown in Fig. 5. The energy consumption per stack and in the whole set-up can be calculated using Eqs. 9 and 10.

The procedure presented here will end up in the same membrane area than using the model presented by Lee et al. [11]. However, by assuming a more realistic current distribution on the different stack stages it is possible to have a good estimation of the number of stack required.

3 Experimental

3.1 ED set-up

The LCD and batch experiments were carried out in a standard lab-unit consisting on 3 vessels with 2 L of capacity, 3 pumps Verder VMD15 (max flow 15 L/min), a rectifier Delta Elektronik SM 7020-D with auto ranging of 0–70 V/0–20 A and two flow and pressure measurements devices. The stack geometry offers a 0.01 m^2 of active membrane area with a channel thickness of 0.003 m in each compartment. The experiments were done in batch recirculation mode with 1.5 L of waste stream in the diluate and concentrate sides. An electrolyte solution (2 L of Na_2SO_4 0.13 M) was recirculated in a close cycle between

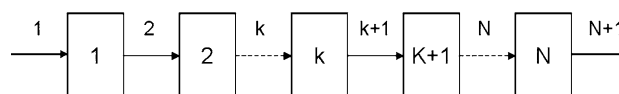
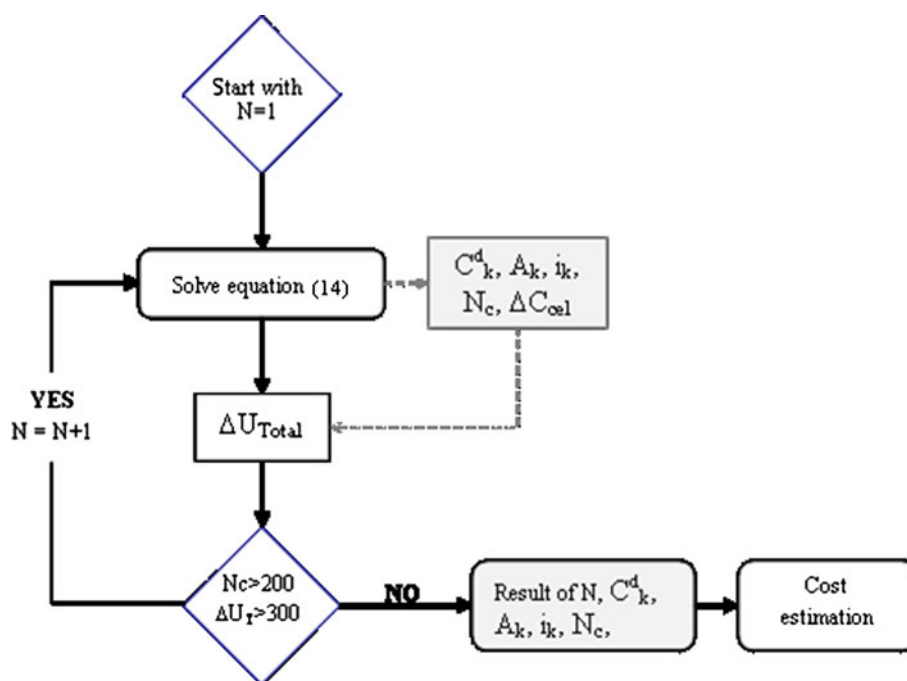


Fig. 4 Stacks in series representing only diluate concentration

Fig. 5 Algorithm to solve the number of stacks required according a variable current density model



the electrodes chambers. The stack was built with 3 cells (4 CEM and 3 AEM). During the experiment, voltage, pH, temperature, volume level, conductivity and concentrations were monitored.

RALEX ion exchange membranes (produced in MEGA a.s. Stráž pod Ralskem, Czech Republic) were used, their main properties are mentioned in Table 1 according to the supplier.

3.2 Case of study

The goal is the treatment of an effluent stream from an anaerobic digester–decanter combination to create a low ion-concentrated stream and a rich or concentrated stream of the minerals for further applications. The main requirement for this stream is a factor of ion depletion that will allow further application of the diluate stream as green-houses irrigation water, which is about a 70 % of ion separation.

Table 1 Mega membrane properties

Property	AEM	CEM
Material		
Ion-exchange groups	R-(CH ₃) ₃ N ⁺	R-SO ₃ ⁻
Binder	Polyethylene	
Thickness (mm)		
Dry	0.45	0.45
Swelled	0.85	0.7
Resistance (Ω cm ²)	<7	<8.5
Counter ion transference number	0.95	0.95
Exchange capacity (mval/g)	1.8	2.2

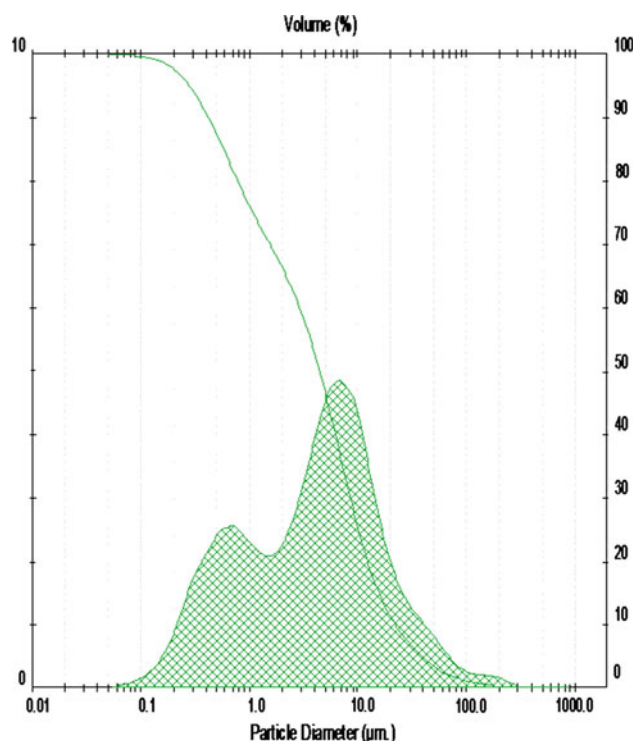
The composition of the stream is presented in Table 2. The components in major concentration are marked as bold letters. This stream contains approximately 9,300 ppm of total dissolved solids (TDS) and suspended solids with a particle size distribution according to Fig. 6. These particles are a mix of unconverted reactants and biological sludge.

Anions (chloride, phosphate, sulphate, nitrate and oxalate) were analysed according to the supplier's application note using an HPLC system (DX-120, Dionex) with conductivity detection (ED40, Dionex). Elution was performed isocratically using a mobile phase composed of 2 mM sodium carbonate and 0.75 mM sodium hydrogen carbonate. The anions were separated using an IonPac AS9-SC carbonate eluent anion exchange column (4.0 × 250 mm²; Dionex) in combination with an IonPacAG9-SC guard column (4.0 × 50 mm²; Dionex), both of which were at room temperature (approximately 20 °C). The flow rate was 1.5 mL/min and the injection volume was 25 μL. The detector was used in conductivity mode (range 30 μS, SRS 100 mA). The data were processed using Empower software [16].

Cations (sodium, potassium, calcium, magnesium and phosphorus) were analysed using inductively coupled plasma with optical emission spectrometry (ICP-OES, iCAP6000 duo, Thermo Scientific, Waltham, MA, USA) after acid digestion of the samples using a microwave accelerated reaction system (MARS 5, CEM, Matthews, NC, USA). Test portions of 0.5 g sample and 10 mL of concentrated nitric acid (65 %, Suprapur, Merck, Darmstadt, Germany) were added to a reaction vessel (Omni,

Table 2 Feed composition

Component	
Dry matter (%)	2.1
Ash content (%)	1.1
HCl-insoluble matter (%)	0.1
pH	8.6
Total N (mg/kg)	1,700
Free N (mg/kg)	1,300
COD (mg/L)	13,800
Chloride (mg/kg)	1,800
Nitrate (mg/kg)	<10
Phosphate (mg/kg)	110
Sulphate (mg/kg)	140
Oxalate (mg/kg)	<10
Sodium (mg/kg)	2,900
Potassium (mg/kg)	1,800
Magnesium (mg/kg)	170
Calcium (mg/kg)	550
Phosphor (mg/kg)	130

**Fig. 6** Particle size distribution

CEM). Acid digestion was carried out during 15 min at a microwave power intensity of 1,600 W and a temperature of 180 °C. Afterwards, the sample solutions were cooled and quantitatively transferred to 100-mL volumetric flasks before analysis. ICP-OES analysis was carried out using

yttrium as internal standard. RF power was set to 1,150 W and pump rate was 50 rpm. The auxiliary and nebulizer gas flow rate were both 0.5 L/min and the purge gas flow was set to normal. The data were processed using iTEVAiCAP software (Thermo Scientific).

4 Results

4.1 LCD results

The LCD values have been obtained following the procedure described in Sect. 2.1. In this case, the same velocity was used in all the experiments, therefore the limiting value has been calculated at different dilutions of the original sample. There are more details of LCD results in the Supplementary information of this article.

The final results are represented in Fig. 7 in which the LCD values are represented as function of Na^+ concentration which is of one majority components, represented in Eq. 15, where i_{lim} and C units are A/m^2 and g/L , respectively. The LCD values have been used for the selection of the voltage applied in the batch experiment.

$$i_{\text{lim}} = 91.05C^{0.30}. \quad (15)$$

4.2 Batch experiment results

A batch experiment was performed according to the description in Sect. 3.1. The experiment was run until the conductivity on the diluate side was less than 30 % of the initial value. The operation mode was at fixed voltage obtaining current density below the LCD values, according to Table 3.

In the table, the presented values are absolutes at the time indicated in the head row of the table. The current density was decreasing as long as the separation took place because the resistance increases in the diluate side, these values were considerably lower than the LCD obtained with Eq. 15. The recirculation flow was 80 and 130 L/h for diluate and concentrate, respectively, which represents 7 and 12 cm/s.

The evolution of the conductivity on both compartments is represented in Fig. 8 as a function of the time. The pH changes during the experiments were measured with an average value for every compartment of:

- Diluate: 7.6 ± 0.7 .
- Concentrate: 6.9 ± 0.6 .
- Electrolyte: 7.5 ± 0.8 .

From the experimental results is also possible to determine the resistance evolution during the experiment on two contributions: wet membrane resistance (and may be fouled) and the concentrate and diluted compartments

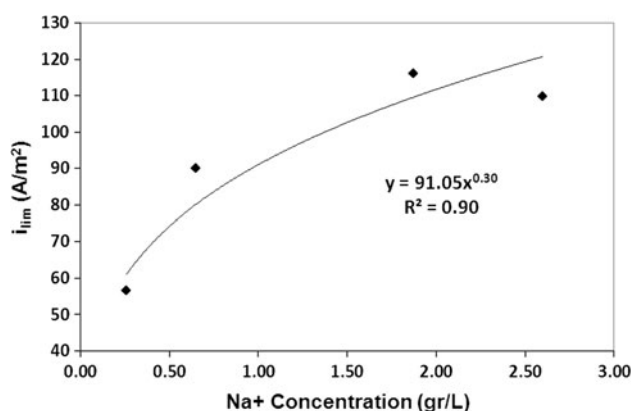


Fig. 7 LCD correlation as function of Na^+ concentration

solutions. The global voltage over the stack can be theoretically calculated by adding Eqs. 7 and 8 for the number of cells used (3 in this case):

$$(\Delta U_{\text{stack}} - \Delta U_{\text{ov}}) = N(\Delta U_{\text{sol}} + \Delta U_{\text{mem}}) \quad (16)$$

The membranes resistance can be obtained rearranging Eq. 16, where 1.3 V was assumed as overvoltage potential, ΔU_{ov} . The overvoltage should be measured for a better estimation. However, the effect of overvoltage on industrial installations is normally negligible [11]. The average resistance values obtained for a single membrane is $89 \, \Omega \, \text{cm}^2$ at the beginning of the experiment. It was also observed that most of the resistance is on the wet membranes, these are typical results at low concentration when direct voltage is measured. A full explanation of the phenomena was described in the study of Długolecki et al. [17] and it was also found that membrane resistance is dependent of solution concentration at lower concentrations than the typically used by membrane suppliers (0.5–1 M NaCl). By direct voltage measurement, the limiting layers are included in the resistance calculation and the resistance in the diluate layer will increase due to ion depletion [18].

In Fig. 9, the single membrane resistance is represented along the experiment. There are also some reasons for the membrane resistance increase in time:

- As the membrane is wet from both sides with different solutions conductivities, the solution inside will have an intermediate value which is practically unknown (see Fig. 3), but in overall it decreases and resistance increases.

- The particles present in the sample where gradually been accumulated in the compartments forming a cake that was easily removed after disassembling the stack. There was also measured a decrease in the flow through the compartments, 22 and 27 % of flow reduction in diluate and concentrate compartments, respectively.
- Fouling with organic material is the other important contribution. It is known that large charged molecules (e.g. humic substances [19]) can block ion channels within anion exchange membranes. No scaling was visually detected.

4.3 Ion separation analysis

In multi-component separation with ED, it is not straightforward to predict which ions will be preferentially transported, some reasons are: ion competition [20, 21], monovalent ion selectivity of the membrane and pH influence [22]. The analysis of ion concentrations at the beginning and end of the batch test is shown in Table 4 together with percentage of separation and averaged current efficiency obtained with Eq. 6, it was called ‘averaged’ because it only considers the amount of ions transferred from beginning to end of the experiment. The ion concentration was determined according to techniques described in Sect. 3.2.

It is possible to see how the cations Na^+ , NH_4^+ and K^+ with the highest concentration have been removed preferentially; they are all monovalent with preferential transport through the membrane. The rather low transport of calcium and magnesium through the membrane is beneficial to minimise scaling risk in the presence of bicarbonate.

The monovalent anion Cl^- was also highly separated. Moreover, the fact that phosphate has a separation rate higher than sulphate indicates that it has been transported through the membrane in the monovalent form (H_2PO_4^-) which is highly possible at around neutral pH, while sulphate it is mainly in the divalent form (SO_4^{2-}).

The addition of all cation efficiencies (last column of Table 4) is about 86 % respect to the current applied. This efficiency should match the corresponding one obtained from anions addition, that is not the case in this table because volatile fatty acids (VFAs) and HCO_3^- have not been accounted (analysed); however, they are present in these wastewater according with other tests performed.

Table 3 Comparison between current applied and LCD

Time (min)	90	150	210	325	365
Voltage (V)	5	5	5	10	22
Current density (A/m^2)	60	50	40	30	40
LCD (A/m^2) Eq. 15	127	–	115	–	98

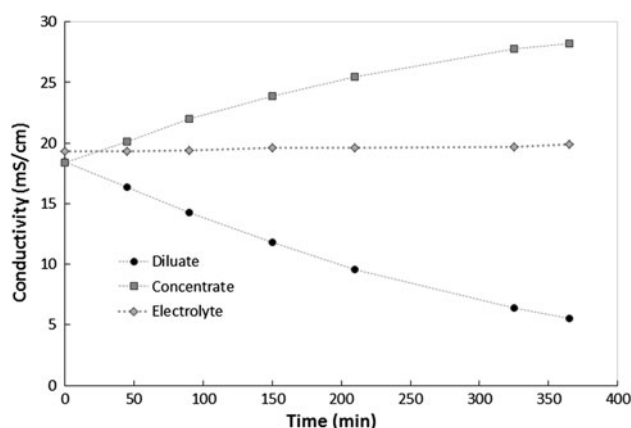


Fig. 8 Conductivity evolution in the batch experiment

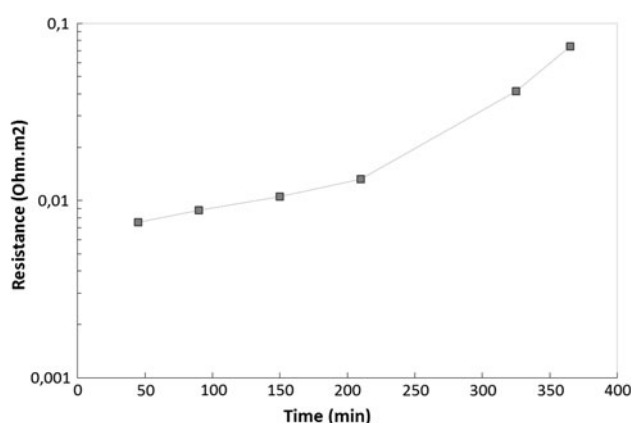


Fig. 9 Stack, membranes and solutions resistance during the batch experiment

4.4 Theoretical design

A theoretical design can be assessed using the experimental results coming from the laboratory unit and the model described in Sect. 2. Particularly, the case under study has been designed for a continuous operation using the combination of the models presented on this article. For every individual stack, the model described in Sect. 2.6 has been used and the algorithm presented on Sect. 3 was used to obtain the current density, number of stacks and ions separation on every unit.

The operational conditions and the experimental parameters obtained from the lab results are listed in Table 5. The water reclamation parameter has been defined as the amount of water recovered after the treatment in comparison to the throughput. In ED, it has been calculated as the flow ratio between diluate and feed streams. In normal ED set-ups, the spacer thickness is about 0.5–1 mm [23], in this case due to the presence of small particles a thicker channel was used for the experiments as well as for the design.

Membrane resistance was obtained from Fig. 9 assuming that CEM and AEM have the same electric resistance.

Current efficiency was calculated for every ion in the solution according to Table 4 and for design purpose the ion targeted for separation is Na^+ and the corresponding efficiency is about 40 %.

In Table 6, the main results of the design are presented for water reclamation of 75 %. Moreover, in Fig. 10, the diluate and concentrate concentration is represented at the exit value of every stack. In this figure, the applied current density is also represented in the right vertical axis. In Fig. 11, a benchmarking of membrane area and energy necessary for different water reclamation rates is represented. The amount of stacks required for such of separation was included on the number next to each point. From this figure, the energy required for the particular case was calculated for a continuous operation in a range from 6 to 11 kWh/m³ of feed stream. The TDSs in the stream is about 10,000 ppm which is in the range of brackish water ranging from 1,500 to 15,000 ppm, therefore the energy consumption obtained in this study can be compared with other technologies presented for brackish water desalination. Younos and Tulou [24] have presented energy consumption values of 1.7 and 4.2–6.4 kWh/m³ for ED and reverse osmosis, respectively. Plappally and Lienhard [1] reported energy consumption depending on the TDS content: 500, 1000 and 4000 ppm; TDS consumes 0.66, 0.79 and 1.59 kWh/m³, respectively, which when extrapolated to 10,000 ppm is around 3.2 kWh/m³. The estimation made for the industrial wastewater only based on lab experiments is higher than previous references. It could be explained by the high resistance values obtained from the lab experiments as consequence of the suspended material present in the sample and dissolved organics on anion exchange membranes. Moreover, the spacer thickness used in this study is at least three times thicker than normal spacers which indeed increase the resistance of the compartments.

5 Conclusions

A first estimate design for an industrial wastewater desalination plant has been done for the effluent stream from a digester–decanter system. The composition of the stream includes several anions and cations as well as organic material and suspended mater. In such a complex stream, it is very complex to predict the ions separation and performance of an ED installation. Therefore, lab experiments were carried out to elucidate the most relevant design parameters like LCD, membrane resistance, ions separation rate and current efficiency.

Table 4 Ion analysis and separation rate after 6 h

	Initial (mg)	Final (mg)	Separation (%)	Averaged efficiency (%)
Anions				
Cl ⁻	2,700	116.7	95.7	22.1
SO ₄ ⁻²	210	194.6	7.4	0.0
PO ₄ ⁻³	110	90.0	29.3	0.2
Cations				
Na ⁺	4,350	1,284	70.5	40.5
K ⁺	2,700	389.1	85.6	18.0
NH ₄ ⁺	1,950	479.9	75.4	24.8
Mg ⁺²	255	127.1	50.2	1.6
Ca ⁺²	825	687.4	16.7	1.0

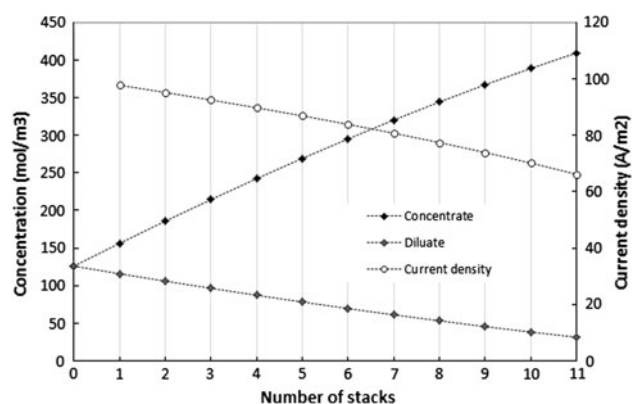
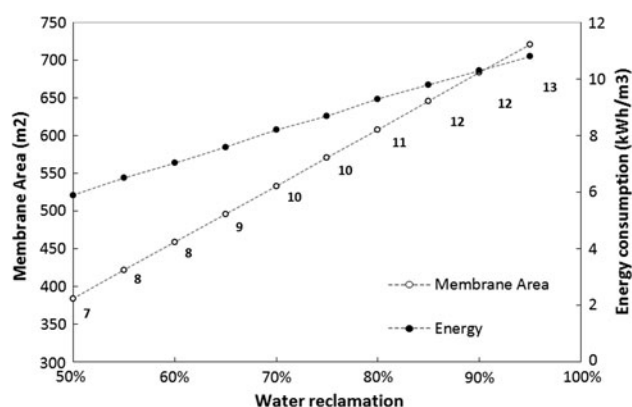
Table 5 Fix parameter for ED design

Operation	
Na ⁺ recovery rate (eq/s)	0.75
Feed flow (m ³ /h)	10
Water reclamation (%)	75
Spacer thickness (m)	0.003
Width (m)	0.4
Height (m)	1
Experimental	
Limiting current density (A/m ²)	Eq. 15
Membrane resistance (Ω cm ²)	89
Current efficiency	0.4

Table 6 Model design results for continuous operation

Model results	Values
Membrane area (m ²)	570
Number of cells	1,425
Number of stacks	9
Energy consumption (kWh/m ³)	8.7

The results show that it is possible to separate mono-valent ions, Cl⁻, K⁺ and Na⁺ at high percentages of about 96, 86 and 70 %, respectively, while the divalent, SO₄⁻², HPO₄⁻², Mg⁺² and Ca⁺² ions were less separated (less than 50 %). In order to perform the design of the ED process, Na⁺ ion was selected as target ion and a recovery of 75 % of Na⁺ will ensure a higher depletion of the other mono-valent ions according to the lab batch experiments. The resistance and current efficiency were also assessed from the batch experiments. The energy required for the particular case was calculated for a continuous operation in a range from 6 to 11 kWh/m³ of feed stream depending on

**Fig. 10** Diluate and concentrate concentration at the exit of every stack**Fig. 11** Benchmarking of membrane area and energy required for different water reclamation rates. The numbers next to the points represent the amount of stacks required

the water reclamation rate which is comparable with the data presented in literature for brackish water desalination. For those conditions, the number of stack required varies between 7 and 13 with membrane area requirement of between 380 and 720 m². The next step to validate this study is been carried out in a pilot scale plant for long-term operation.

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